(51) International Patent Classification 5 :		(11) International Publication Number:	WO 90/12076
C10G 11/18	**A1**	(43) International Publication Date:	18 October 1990 (18.10.90)

(21) International Application Number: PCT/US90/01881

(22) International Filing Date: 6 April 1990 (06.04.90)

(30) Priority data:
335,642   10 April 1989 (10.04.89)   US

(71) Applicant: MOBIL OIL CORPORATION [US/US]; 150 East 42nd Street, New York, NY 10017 (US).

(72) Inventor: OWEN, Hartley ; 5 Riverview Terrace, Belle Mead, NJ 08502 (US).

(74) Agents: SUNG, Tak, K. et al.; Mobil Oil Corporation, 3225 Gallows Road, Fairfax, VA 22037 (US).

(81) Designated States: AT (European patent), AU, BE (European patent), CA, CH (European patent), DE (European patent), DK (European patent), ES (European patent), FR (European patent), GB (European patent), IT (European patent), JP, LU (European patent), NL (European patent), SE (European patent).

Published
With international search report.

(54) Title: HEAVY OIL CATALYTIC CRACKING PROCESS AND APPARATUS

(57) Abstract

A fluidized catalytic cracking process operates with a hot stripper to improve stripping of spent catalyst from the FCC process. The catalyst from the hot stripper is cooled by direct contact heat exchange with a source or cooled regenerated catalyst. Cooled catalyst may contact hot, stripped catalyst in the base of the stripper or downstream of the stripper. The cooled, stripped catalyst has reduced hydrogen, sulfur and coke content, improves regeneration efficiency, and reduces hydrothermal degradation of catalyst.
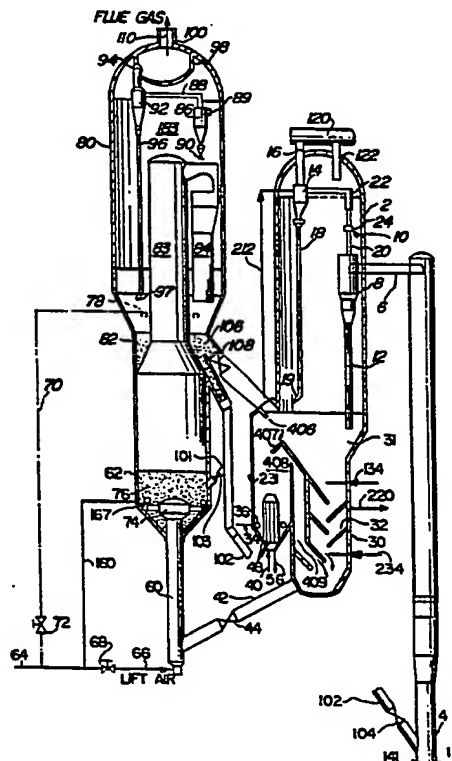
DESIGNATIONS OF "DE"

Until further notice, any designation of "DE" in any international application whose international filing date is prior to October 3, 1990, shall have effect in the territory of the Federal Republic of Germany with the exception of the territory of the former German Democratic Republic.

HEAVY OIL CATALYTIC CRACKING PROCESS AND APPARATUS

This invention relates to regeneration of coked cracking catalyst in a fluidized bed.

Catalytic cracking is the backbone of many refineries.
5   It converts heavy feeds into lighter products by catalytically cracking large molecules into smaller molecules. Catalytic cracking operates at low pressures, without hydrogen addition, in contrast to hydrocracking, which operates at high hydrogen partial pressures. Catalytic cracking is inherently safe as it
10   operates with very little oil actually in inventory during the cracking process.

There are two main variants of the catalytic cracking process: moving bed and the far more popular and efficient fluidized bed process.

15   In the fluidized catalytic cracking (FCC) process, catalyst, having a particle size and color resembling table salt and pepper, circulates between a cracking reactor and a catalyst regenerator. In the reactor, hydrocarbon feed contacts a source of hot, regenerated catalyst. The hot catalyst vaporizes and
20   cracks the feed at 425-600°C, usually 460-560°C. The cracking reaction deposits carbonaceous hydrocarbons or coke on the catalyst, thereby deactivating the catalyst. The cracked products are separated from the coked catalyst. The coked catalyst is stripped of volatiles, usually with steam, in a
25   catalyst stripper and the stripped catalyst is then regenerated. The catalyst regenerator burns coke from the catalyst with oxygen containing gas, usually air. Decoking restores catalyst activity and simultaneously heats the catalyst to, e.g., 500-900°C, usually 600-750°C. This heated catalyst is recycled to the
30   cracking reactor to crack more fresh feed. Flue gas formed by burning coke in the regenerator may be treated for removal of

—2—

particulates and for conversion of carbon monoxide, after which the flue gas is normally discharged into the atmosphere.

Catalytic cracking is endothermic, i.e., it consumes heat. The heat for cracking is supplied at first by the hot regenerated catalyst from the regenerator. Ultimately, it is the feed which supplies the heat needed to crack the feed. Some of the feed deposits as coke on the catalyst, and the burning of this coke generates heat in the regenerator, which is recycled to the reactor in the form of hot catalyst.

Catalytic cracking has undergone progressive development since the 1940s. The trend of development of the fluid catalytic cracking (FCC) process has been to all riser cracking and use of zeolite catalysts.

Riser cracking gives higher yields of valuable products than dense bed cracking. Most FCC units now use all riser cracking, with hydrocarbon residence times in the riser of less than 10 seconds, and even less than 5 seconds.

Zeolite-containing catalysts having high activity and selectivity are now used in most FCC units. These catalysts work best when the coke content on the catalyst after regeneration is less than 0.1 wt %, and preferably less than 0.05 wt %.

To regenerate FCC catalysts to these low residual carbon levels, and to burn CO completely to CO_2 within the regenerator (to conserve heat and minimize air pollution) many FCC operators add a CO combustion promoter metal to the catalyst or to the regenerator.

As the process and catalyst improved, refiners attempted to use the process to upgrade a wider range of feedstocks, in particular, feedstocks that were heavier, and also contained more metals and sulfur than had previously been permitted in the feed to a fluid catalytic cracking unit.

These heavier, dirtier feeds have placed a growing demand on the regenerator. Processing resids has exacerbated four

—3—

existing problem areas in the regenerator, sulfur, steam, temperature and NO_x. These problems will each be reviewed in more detail below.

SULFUR

5   Much of the sulfur in the feed ends up as SO_x in the regenerator flue gas. Higher sulfur levels in the feed, combined with a more complete regeneration of the catalyst in the regenerator increases the amount of SO_x in the regenerator flue gas. Some attempts have been made to minimize the amount of SO_x

10   discharged to the atmosphere through the flue gas by including catalyst additives or agents to react with the SO_x in the flue gas. These agents pass with the regenerated catalyst back to the FCC reactor where the reducing atmosphere releases the sulfur compounds as H_2S.

15   Unfortunately, the conditions in most FCC regenerators are not the best for SO_x adsorption. The high temperatures in modern FCC regenerators (up to 870°C (1600°F)) impair SO_x adsorption. One way to minimize SO_x in flue gas is to pass catalyst from the FCC reactor to a long residence time steam

20   stripper, as disclosed in U.S. Patent No. 4,481,103 to Krambeck et al. This process preferably steam strips spent catalyst at 500-550°C (932 to 1022°F), which is beneficial but not sufficient to remove some undesirable sulfur- or hydrogen-containing components.

25   ### STEAM

Steam is always present in FCC regenerators although it is known to cause catalyst deactivation. Steam is not intentionally added, but is invariably present, usually as adsorbed or entrained steam from steam stripping or catalyst or

30   as water of combustion formed in the regenerator.

—4—

Poor stripping leads to a double dose of steam in the regenerator, first from the adsorbed or entrained steam and second from hydrocarbons left on the catalyst due to poor catalyst stripping. Catalyst passing from an FCC stripper to an FCC regenerator contains hydrogen-containing components, such as coke or unstripped hydrocarbons adhering thereto. This hydrogen burns in the regenerator to form water and cause hydrothermal degradation.

U.S. Patent No. 4,336,160 to Dean et al attempts to reduce hydrothermal degradation by staged regeneration. However, the flue gas from both stages of regeneration contains SO_x which is difficult to clean. It would be beneficial, even in staged regeneration, if the amount of water precursors present on stripped catalyst was reduced.

Steaming of catalyst becomes more of a problem as regenerators get hotter. Higher temperatures greatly accelerate the deactivating effects of steam.

TEMPERATURE

Regenerators are operating at higher and higher temperatures. This is because most FCC units are heat balanced, that is, the endothermic heat of the cracking reaction is supplied by burning the coke deposited on the catalyst. With heavier feeds, more coke is deposited on the catalyst than is needed for the cracking reaction. The regenerator gets hotter, and the extra heat is rejected as high temperature flue gas. Many refiners severely limit the amount of resid or similar high Conradson Carbon Residue (CCR) feeds to that amount which can be tolerated by the unit. High temperatures are a problem for the metallurgy of many units, but more importantly, are a problem for the catalyst. In the regenerator, the burning of coke and unstripped hydrocarbons leads to much higher surface temperatures on the catalyst than the measured dense bed or dilute phase

—5—

temperature. This is discussed by Occelli et al in Dual-Function Cracking Catalyst Mixtures, Ch. 12, Fluid Catalytic Cracking, ACS Symposium Series 375, American Chemical Society, Washington, D.C., 1988.

5   Some regenerator temperature control is possible by adjusting the CO/CO_2 ratio produced in the regenerator. Burning coke partially to CO produces less heat than complete combustion to CO_2. However, in some cases, this control is insufficient, and also leads to increased CO emissions, which can be a problem

10   unless a CO boiler is present.

        U.S. Patent No. 4,353,812 to Lomas et al discloses cooling catalyst from a regenerator by passing it through the shell side of a heat-exchanger with a cooling medium through the tube side. The cooled catalyst is recycled to the regeneration

15   zone. This approach removes heat from the regenerator, but does not prevent poorly, or even well, stripped catalyst from experiencing very high surface or localized temperatures in the regenerator. The Lomas process does not control the temperature of catalyst from the reactor stripper to the regenerator.

20   The prior art also used dense or dilute phase regenerated fluid catalyst heat removal zones or heat-exchangers that are remote from, and external to, the regenerator vessel to cool hot regenerated catalyst for return to the regenerator. Examples of such processes are found in U.S. Patent Nos. 2,970,117 to Harper;

25   2,873,175 to Owens; 2,862,798 to McKinney; 2,596,748 to Watson et al; 2,515,156 to Jahnig et al; 2,492,948 to Berger; and 2,506,123 to Watson. In these processes the regenerator operating temperature is affected by the temperature of catalyst from the stripper.

30   NO_x

        Burning of nitrogenous compounds in FCC regenerators has long led to creation of minor amounts of NO_x, some of which were

emitted with the regenerator flue gas. Usually these emissions
were not much of a problem because of relatively low temperature,
a relatively reducing atmosphere from partial combustion of CO
and the absence of catalytic metals like Pt in the regenerator
5   which increase NO_x production.

Many FCC units now operate at higher temperatures, with a
more oxidizing atmosphere, and use CO combustion promoters such
as Pt. These changes in regenerator operation reduce CO
emissions, but usually increase nitrogen oxides (NO_x) in the
10 regenerator flue gas. It is difficult in a catalyst regenerator
to completely burn coke and CO in the regenerator without
increasing the NO_x content of the regenerator flue gas, so NO_x
emissions are now frequently a problem.

To reduce NO_x emissions, it has been suggested to use
15 combustion promoters, steam treatment of conventinoal metallic CO
combustion promoter, multi-stage FCC regenerators, countercurrent
regeneration, addition of a vaporizable fuel to the upper portion
of a FCC regenerator, adjust the concentration of CO combustion
promoter and reduce the amount of flue gas by using oxygen rather
20 than air. These approaches still may fail to meet the ever more
stringent NO_x emissions limits set by local governing bodies.
Much of the NO_x formed is not the result of combustion of N_2
within the FCC regenerator, but rather combustion of
nitrogen-containing compounds in the coke entering the FCC
25 regenerator. Bi-metallic combustion promoters are probably best
at minimizing NO_x formation from N_2.

Unfortunately, the trend to heavier feeds usually means
that the amount of nitrogen compounds on the coke will increase
and that NO_x emissions will increase. Higher regenerator
30 temperatures also tend to increase NO_x emissions. It would be
beneficial, in many refineries, to have a way to burn at least a
large portion of the nitrogenous coke in a relatively reducing
atmosphere, so that much of the NO_x formed could be converted

—7—

into N_2 within the regenerator. Unfortunately, most existing regenerator designs cannot operate efficiently at such conditions, i.e., with a reducing atmosphere.

5   It would be beneficial if a better stripping process were available which would permit increased recovery of valuable, strippable hydrocarbons. There is a need for a higher temperature stripper, which will not lead to a higher temperature regenerator. There is a special need to remove more hydrogen from spent catalyst to minimize hydrothermal degradation in the
10   regenerator. It would be further advantageous to remove more sulfur-containing compounds from spent catalyst prior to regeneration to minimize SO_x in the regenerator flue gas. Also, it would be advantageous to have a better way to control regenerator temperature.

15   The present invention provides a way to achieve much better high temperature stripping of coked FCC catalyst. The present invention not only improves stripping, and increases the yield of valuable liquid product, it reduces the load placed on the catalyst regenerator, minimizes SO_x emissions, and permits
20   the unit to process more difficult feeds. Regenerator temperatures can be reduced, or maintained constant while processing worse feeds, and the amount of hydrothermal deactivation of catalyst in the regenerator can be reduced.

According to the present invention, a fluidized catalytic
25   cracking process is provided wherein a heavy hydrocarbon feed comprising hydrocarbons having a boiling point above 343°C (650°F) is catalytically cracked to lighter products comprising the steps of: catalytically cracking the feed in a catalytic cracking zone operating at catalytic cracking conditions by
30   contacting the feed with a source of hot regenerated catalyst to produce a cracking zone effluent mixture having an effluent temperature and comprising cracked products and spent cracking catalyst containing coke and strippable hydrocarbons; separating

the cracking zone effluent mixture into a cracked product rich
vapor phase and a solids rich phase comprising the spent catalyst
and strippable hydrocarbons, the solids rich phase having a
temperature; heating the solids rich phase by mixing it with a
5   source of hot regenerated catalyst having a higher temperature
than the solids rich phase to produce a catalyst mixture
comprising spent and regenerated catalyst having a catalyst
mixture temperature intermediate the solids rich phase
temperature and the temperature of the regenerated catalyst;
10 stripping in a primary stripping stage the catalyst mixture with
a stripping gas to remove strippable compounds from spent
catalyst to produce a stripped catalyst stream; cooling a source
of hot regenerated catalyst by passing hot regenerated catalyst
through a cooling means to produce cooled regenerated catalyst;
15 cooling the stripped catalyst stream by direct contact heat
exchange with cooled regenerated catalyst to produce a cooled,
stripped catalyst stream; regenerating the cooled, stripped
catalyst stream by contact with oxygen or an oxygen containing
gas in a regenerating means to produce hot regenerated catalyst
20 as a result of combustion of coke on the spent catalyst;
recycling to the cracking reaction zone a portion of the hot
regenerated catalyst to crack more hydrocarbon feed; recycling
to the primary stripping stage a portion of the regenerated
catalyst to heat spent catalyst, and recycling to the regenerated
25 catalyst cooling means a portion of the regenerated catalyst to
produce cooled regenerated catalyst.

       In another embodiment, the present invention provides an
apparatus for the fluidized catalytic cracking of a heavy
hydrocarbon feed comprising hydrocarbons having a boiling point
30 above 343°F (650°F) to lighter products by contacting the feed
with catalytic cracking catalyst comprising a catalytic cracking
reactor means having an inlet connective with the feed and with a
source of hot regenerated catalyst and having an outlet for

—9—

discharging a cracking zone effluent mixture comprising cracked products and spent cracking catalyst containing coke and strippable hydrocarbons; a separation means connective with the reactor outlet for separating the cracking zone effluent mixture
5   into a cracked product rich vapor phase and a solids rich phase comprising the spent catalyst and strippable hydrocarbons; a hot stripping means having an upper portion and a lower portion and comprising an inlet for a source of hot regenerated cracking catalyst in the upper portion thereof, an inlet for spent
10   catalyst, an inlet for a stripping gas, a stripping vapor outlet for stripping vapors and a solids outlet for discharge of hot stripped solids in a lower portion thereof; a regenerated catalyst cooling means comprising a vessel adapted to contain a fluidized bed of catalyst and having an inlet connective with a
15   source of hot regenerated catalyst, a heat exchange means immersed at an elevation within the fluidized bed of catalyst for removal of heat to produce cooled regenerated catalyst, an inlet for a fluidizing gas , and an outlet for cooled, regenerated catalyst; a direct contact heat exchange means for contact and
20   cooling of hot stripped solids with cooled regenerated catalyst to produce cooled stripped catalyst; a catalyst regeneration means having an inlet connective with the cooled, stripped catalyst, a regeneration gas inlet, a flue gas outlet, and an outlet for removal of hot regenerated catalyst; and catalyst
25   recycle means connective with the catalytic cracking reaction zone, the primary stripping zone, and the hot regenerated catalyst cooling means.

The Figure is a simplified schematic view of an FCC unit with a hot stripper of the invention.
30   The present invention can be better understood by reviewing it in conjunction with the Figure, which illustrates a fluid catalytic cracking system of the present invention.

—10—

Although a preferred FCC unit is shown, any riser reactor and
regenerator can be used in the present invention.

A heavy feed is charged via line 1 to the lower end of a
riser cracking FCC reactor 4. Hot regenerated catalyst is added
5   via standpipe 102 and control valve 104 to mix with the feed.
Preferably, some atomizing steam is added via line 141 to the
base of the riser, usually with the feed . With heavier feeds,
e. g. , a resid, 2-10 wt.% steam may be used. A
hydrocarbon-catalyst mixture rises as a generally dilute phase
10 through riser 4. Cracked products and coked catalyst are
discharged via riser effluent conduit 6 into first stage cyclone
8 in vessel 2. The riser top temperature, the temperature in
conduit 6, ranges between 480 and 615°C (900 and 1150°F), and
preferably between 538 and 595°C (1000 and 1050°F). The riser
15 top temperature is usually controlled by adjusting the catalyst
to oil ratio in riser 4 or by varying feed preheat.

Cyclone 8 separates most of the catalyst from the cracked
products and discharges this catalyst down via dipleg 12 to a
stripping zone 30 located in a lower portion of vessel 2. Vapor
20 and minor amounts of catalyst exit cyclone 8 via gas effluent
conduit 20 and flow into connector 24, which allows for thermal
expansion, to conduit 22 which leads to a second stage reactor
cyclone 14. The second cyclone 14 recovers some additional
catalyst which is discharged via dipleg 18 to the stripping zone
25 30.

The second stage cyclone overhead stream, which includes
cracked products and catalyst fines, passes via effluent conduit
16 and line 120 to product fractionators not shown in the figure.
Stripping vapors enter the atmosphere of the vessel 2 and exit
30 this vessel via outlet line 22 or by passing through the annular
space 10 defined by outlet 20 and inlet 24.

The coked catalyst discharged from the cyclone diplegs
collects as a bed of catalyst 31 in the stripping zone 30.

—11—

Dipleg 12 is sealed by being extended into the catalyst bed 31. Dipleg 18 is sealed by a trickle valve 19.

Although only two cyclones 8 and 14 are shown, many cyclones, 4 to 8, are usually used in each cyclone separation stage. A preferred closed cyclone system is described in U.S.
5   Patent No. 4,502,947 to Haddad et al.

Stripper 30 provides for "hot stripping" in bed 31. Spent catalyst is mixed in bed 31 with hot catalyst from the regenerator. Direct contact heat exchange heats spent catalyst. The regenerated catalyst, which has a temperature from 55°C
10 (100°F) above the stripping zone 30 to 871°C (1600°F), heats spent catalyst in bed 31. Catalyst from regenerator 80 enters vessel 2 via transfer line 106, and slide valve 108 which controls catalyst flow. Adding hot, regenerated catalyst permits first stage stripping at from 55°C (100°F) above the riser
15 reactor outlet temperature and 816°C (1500°F). Preferably, the first stage stripping zone operates at least 83°C (150°F) above the riser top temperature, but below 760°C (1400°F).

In bed 31 a stripping gas, preferably steam, flows countercurrent to the catalyst. The stripping gas is preferably
20 introduced into a lower portion of bed 31 by one or more conduits 134. Bed 31 preferably contains trays or baffles 32. The trays may be disc- and doughnut-shaped and may be perforated or unperforated.

Stripping zone 31 may contain an additional point or
25 points of steam or other stripping gas injection at lower points in the bed, such as by line 234 in the base of the stripping zone. The stripping gas added at the base, such as 234, may be added primarily to promote better fluidization as the base of the stripper and perform little stripping, thus an entirely different
30 stripping gas may be used, such as flue gas. Multiple points of withdrawal of stripping vapor, as by exhaust line 220, may be provided.

—12—

The spent catalyst residence time in bed 31 in the
stripping zone 30 preferably ranges from 1 to 7 minutes. The
vapor residence time in bed 31 preferably ranges from 0.5 to 30
seconds, and most preferably 0.5 to 5 seconds.

5   High temperature stripping removes coke, sulfur and
hydrogen from the spent catalyst. Coke is removed because carbon
in the unstripped hydrocarbons is burned as coke in the
regenerator. The sulfur is removed as hydrogen sulfide and
mercaptans. The hydrogen is removed as molecular hydrogen,
10   hydrocarbons, and hydrogen sulfide. The removed materials also
increase the recovery of valuable liquid products, because the
stripper vapors can be sent to product recovery with the bulk of
the cracked products from the riser reactor. High temperature
stripping can reduce coke load to the regenerator by 30 to 50% or
15   more and remove 50-80% of the hydrogen as molecular hydrogen,
light hydrocarbons and other hydrogen-containing compounds, and
remove 35 to 55% of the sulfur as hydrogen sulfide and
mercaptans, as well as a portion of nitrogen as ammonia and
cyanides.

20   After high temperature stripping in bed 31, the catalyst
has a much reduced content of strippable hydrocarbons, but is too
hot to be charged to the regenerator. The combination of high
initial temperature, and rapid combustion of residual strippable
hydrocarbons, and to a lesser extent of coke, could result in
25   extremely high localized temperatures on the surface of the
catalyst during regeneration. To reduce the bulk temperature of
the hot stripped catalyst, the present invention provides for
direct contact cooling of catalyst after catalyst stripping.

          The hot stripped catalyst from bed 31 passes down through
30   baffles 32 and is cooled by direct contact heat exchange with
cooled, regenerated catalyst. Opening 406 allows hot,
regenerated catalyst to flow into catalyst cooler 231. A stab in
heat exchanger or tube bundle 48 is inserted into the lower

—13—

portion of bed 231. For effective heat exchange, the bed 231 should be fluidized with a gas or vapor, added via line 34 and distributing means 36. Preferably, steam is not used here, because the freshly regenerated catalyst is very hot, and steam

5 addition would cause unnecessary steaming.

Fluidizing gas 34 not only improves heat transfer across tube bundle 48, it provides a good way to control the amount of catalyst that is cooled, for direct contact cooling, versus the amount of catalyst that is added hot to the stripper, for direct

10 contact heating. When little or no fluidizing gas is added to vessel 231, it fills with catalyst from the regenerator but does not flow out readily. Fluidizing gas expands and fluidizes the bed, permitting it to flow like a liquid through opening 406, down around baffle 407 and back up through opening 408 and

15 through downcomer 409 to contact hot, stripped catalyst in the base of the stripper 30.

Valve 108 controls the total amount of regenerated catalyst sent to the stripper 31. The amount of fluidizing gas determines the split between regenerated catalyst that is added

20 hot, and regenerated catalyst that is added cold, by flowing through heat exchanger section 231.

Although not shown in the drawing, additional stages of baffling, or of stripping may be present downstream of the point of addition of cooled, regenerated catalyst. Line 42 may contain

25 one or more splitters or flow dividers, to promote mixing cooled regenerated catalyst with hot stripped catalyst.

The amount of fluidizing gas added via line 34 also permits some control of the heat transfer coefficient across tube bundle 48, permitting some control of heat transfer from hot

30 catalyst to fluid in line 40 (typically boiler feed water or low grade stream) to produce heated heat transfer fluid in line 56 (typically high grade steam.)

—14—

Preferably the catalyst exiting the stripper is at least 28°C (50°F) cooler than the catalyst in the hot stripper, or bed 31. More preferably, the catalyst leaving the stripper via line 42 is 42 to 111°C (75-200°F) cooler than the catalyst in bed 31.

Stripped cooled catalyst passes via effluent line 42 and valve 44 to the regenerator. A catalyst cooler, not shown, may be provided to further cool the catalyst, if necessary to maintain the regenerator 80 at a temperature between 55°C (100°F) above the temperature of the stripping zone 30 and 871°C (1600°F).

When an external catalyst cooler is used it preferably is an indirect heat-exchanger using a heat-exchange medium such as liquid water (boiler feed water).

The cooled catalyst passes through the conduit 42 into regenerator riser 60. Air and cooled catalyst combine and pass up through an air catalyst disperser 74 into coke combustor 62 in regenerator 80. In bed 62, combustible materials, such as coke on the cooled catalyst, are burned by contact with air or oxygen containing gas. At least a portion of the air passes via line 66 and line 68 to riser-mixer 60.

Preferably the amount of air or oxygen containing gas added via line 66, to the base of the riser mixer 60, is restricted to 50-95% of total air addition to the regenerator 80. Restricting the air addition slows down to some extent the rate of carbon burning in the riser mixer, and in the process of the present invention it is the intent to minimize as much as possible the localized high temperature experienced by the catalyst in the regenerator. Limiting the air limits the burning and temperature rise experienced in the riser mixer, and limits the amount of catalyst deactivation that occurs there. It also ensures that most of the water of combustion, and resulting steam, will be formed at the lowest possible temperature.

—15—

Additional air, preferably 5-50 % of total air, is preferably added to the coke combustor via line 160 and air ring 167. In this way the regenerator 80 can be supplied with as much air as desired, and can achieve complete afterburning of CO to CO$_2$, even while burning much of the hydrocarbons at relatively mild, even reducing conditions, in riser mixer 60.

To achieve the high temperatures usually needed for rapid coke combustion, and to promote CO afterburning, the temperature of fast fluidized bed 76 in the coke combustor 62 may be, and preferably is, increased by recycling some hot regenerated catalyst thereto via line 101 and control valve 103.

In coke combustor 62 the combustion air, regardless of whether added via line 66 or 160, fluidizes the catalyst in bed 76, and subsequently transports the catalyst continuously as a dilute phase through the regenerator riser 83. The dilute phase passes upwardly through the riser 83, through a radial arm 84 attached to the riser 83. Catalyst passes down to form a second relatively dense bed of catalyst 82 located within the regenerator 80.

While most of the catalyst passes down through the radial arms 84, the gases and some catalyst pass into the atmosphere or dilute phase region 183 of the regenerator vessel 80. The gas passes through inlet conduit 89 into the first regenerator cyclone 86. Some catalyst is recovered via a first dipleg 90, while remaining catalyst and gas passes via overhead conduit 88 into a second regenerator cyclone 92. The second cyclone 92 recovers more catalyst, and passes it via a second dipleg 96 having a trickle valve 97 to the second dense bed. Flue gas exits via conduit 94 into plenum chamber 98. A flue gas stream 110 exits the plenum via conduit 100.

The hot, regenerated catalyst forms the bed 82, which is substantially hotter than the stripping zone 30. Bed 82 is at least 55°C (100°F) hotter than stripping zone 31, and preferably

—16—

at least 83°C (150°F) hotter. The regenerator temperature is, at
most, 871°C (1600°F) to prevent deactivating the catalyst.

Optionally, air may also be added via line 70, and
control valve 72, to an air header 78 located in dense bed 82

5   Adding combustion air to second dense bed 82 allows some
of the coke combustion to be shifted to the relatively dry
atmosphere of dense bed 82, and minimize hydrothermal degradation
of catalyst. There is an additional benefit, in that the staged
addition of air limits the temperature rise experienced by the

10   catalyst at each stage, and limits somewhat the amount of time
that the catalyst is at high temperature.

Preferably, the amount of air added at each stage (riser
mixer 60, coke combustor 62, transport riser 83, and second dense
bed 82) is monitored and controlled to have as much hydrogen

15   combustion as soon as possible and at the lowest possible
temperature while carbon combustion occurs as late as possible,
and highest temperatures are reserved for the last stage of the
process. In this way, most of the water of combustion, and most
of the extremely high transient temperatures due to burning of

20   poorly stripped hydrocarbon occur in riser mixer 60 where the
catalyst is coolest. The steam formed will cause hydrothermal
degradation of the zeolite, but the temperature will be so low
that activity loss will be minimized. Reserving some of the coke
burning for the second dense bed will limit the highest

25   temperatures to the driest part of the regenerator. The water of
combustion formed in the riser mixer, or in the coke combustor,
will not contact catalyst in the second dense bed 82, because of
the catalyst flue gas separation which occurs exiting the dilute
phase transport riser 83.

30   There are several constraints on the process. If
complete CO combustion is to be achieved, temperatures in the
dilute phase transport riser must be high enough, or the
concentration of CO combustion promoter must be great enough, to

have essentially complete combustion of CO in the transport
riser. Limiting combustion air to the coke combustor or to the
dilute phase transport riser (to shift some coke combustion to
the second dense bed 82) will make it more difficult to get

5   complete CO combustion in the transport riser. Higher levels of
CO combustion promoter will promote the dilute phase burning of
CO in the transport riser while having much less effect on carbon
burning rates in the coke combustor or transport riser.

        If the unit operates in only partial combustion mode, to

10   allow only partial CO combustion, and shift heat generation, to a
CO boiler downstream of the regenerator, then much greater
latitude re air addition at different points in the regenerator
is possible. Partial CO combustion will also greatly reduce
emissions of NO_x associated with the regenerator. Partial CO

15   combustion is a good way to accommodate unusually bad feeds, with
CCR levels exceeding 5 or 10 wt %. Downstream combustion, in a
CO boiler, also allows the coke burning capacity of the
regenerator to increase and permits much more coke to be burned
using an existing air blower of limited capacity

20   Regardless of the relative amounts of combustion that
occur in the various zones of the regenerator, and regardless of
whether complete or only partial CO combustion is achieved, the
catalyst in the second dense bed 82 will be the hottest catalyst,
and will be preferred for use as a source of hot, regenerated

25   catalyst for heating spent, coked catalyst in the catalyst
stripper of the invention. Preferably, hot regenerated catalyst
is withdrawn from dense bed 82 and passed via line 106 and
control valve 108 into dense bed of catalyst 31 in stripper 30.

        Now that the invention has been reviewed in connection

30   with the embodiment shown in the Figure, a more detailed
discussion of the different parts or the process and apparatus of
the present invention follows. Many elements of the present

—18—

invention can be conventional, such as the cracking catalyst, so only a limited discussion of such elements is necessary.

FCC FEED

Any conventional FCC feed can be used. The process of
5   the present invention is especially useful for processing difficult charge stocks, those with high levels of CCR material, exceeding 2, 3, 5 and even 10 wt %CCR. The process, especially when operating in a partial CO combustion mode, tolerates feeds which are relatively high in nitrogen content, and which
10   otherwise might result in unacceptable NO_x emissions in conventional FCC units.

The feeds may range from the typical, such as petroleum distillates or residual stocks, either virgin or partially refined, to the atypical, such as coal oils and shale oils. The
15   feed frequently will contain recycled hydrocarbons, such as light and heavy cycle oils which have already been subjected to cracking.

Preferred feeds are gas oils, vacuum gas oils, atmospheric resids, and vacuum resids. The present invention is
20   most useful when feeds boiling above 343°C (650°F) are used, and preferably when the feed contains 5 wt % or 10 wt % or more of material boiling above 538°C (1000°F).

FCC CATALYST

Any commercially available FCC catalyst may be used. The
25   catalyst can be 100% amorphous, but preferably includes some zeolite in a porous refractory matrix such as silica-alumina, clay, or the like. The zeolite is usually 5-40 wt.% of the catalyst, with the rest being matrix. Conventional zeolites include X and Y zeolites, with ultra stable, or relatively high
30   silica Y zeolites being preferred. Dealuminized Y (DEAL Y) and

—19—

ultrahydrophobic Y (UHP Y) zeolites may be used. The zeolites may be stabilized with Rare Earths, e.g., 0.1 to 10 Wt % RE.

Relatively high silica zeolite containing catalysts are preferred for use in the present invention. They withstand the high temperatures usually associated with complete combustion of CO to CO_2 within the FCC regenerator.

The catalyst inventory may also contain one or more additives, either present as separate additive particles or mixed in with each particle of the cracking catalyst. Additives can be added to enhance octane (shape selective zeolites, i.e., those having a Constraint Index of 1-12, and typified by ZSM-5, and other materials having a similar crystal structure), adsorb SO_x (alumina), remove Ni and V (Mg and Ca oxides).

The FCC catalyst composition, _per se_, forms no part of the present invention.

FCC REACTOR CONDITIONS

Conventional FCC reactor conditions may be used. The reactor may be either a riser cracking unit or dense bed unit or both. Riser cracking is highly preferred. Typical riser cracking reaction conditions include catalyst/oil ratios of 0.5:1 to 15:1 and preferably 3:1 to 8:1, and a catalyst/oil contact time of 0.5-50 seconds, and preferably 1-20 seconds.

The FCC reactor conditions, _per se_, are conventional and form no part of the present invention.

CATALYST STRIPPER/COOLER

Direct contact heating and cooling of catalyst around the catalyst stripper is the essence of the present invention.

Heating of the coked, or spent catalyst is the first step. Direct contact heat exchange of spent catalyst with a source of hot regenerated catalyst is used to efficiently heat spent catalyst.

—20—

Spent catalyst from the reactor, usually at 482° to 621°C (900° to 1150°F) preferably at 510° to 593°C (950 to 1100°F), is charged to the stripping zone of the present invention and contacts hot regenerated catalyst at a temperature of 649°-927°C
5   (1200-1700°F), preferably at 704°-871°C (1300-1600°F). The spent and regenerated catalyst can simply be added to a conventional stripping zone with no special mixing steps taken. The slight fluidizing action of the stripping gas, and the normal amount of stirring of catalyst passing through a conventional stripper will
10 provide enough mixing effect to heat the spent catalyst. Some mixing of spent and regenerated catalyst is preferred, both to promote rapid heating of the spent catalyst and to ensure even distribution of spent catalyst through the stripping zone. Mixing of spent and regenerated catalyst may be promoted by
15 providing some additional fluidizing steam or other stripping gas at or just below the point where the two catalyst streams mix. Splitters, baffles or mechanical agitators may also be used if desired.

The amount of hot regenerated catalyst added to spent
20 catalyst can vary greatly depending on the stripping temperature desired and on the amount of heat to be removed via the stripper heat removal means discussed in more detail below. In general, the weight ratio of regenerated to spent catalyst will be from 1:10 to 10:1, preferably 1:5 to 5:1 and most preferably 1:2 to
25 2:1. High ratios of regenerated to spent catalyst will be used when extremely high stripping efficiency are needed or when large amounts of heat removal are sought in the stripper catalyst cooler. Small ratios will be used when the desired stripping temperature, or stripping efficiency can be achieved with smaller
30 amounts of regenerated catalyst, or when heat removal from the stripper cooler must be limited.

—21—

DIRECT CONTACT COOLING

The process of the present invention provides an efficient, and, readily retrofitted, means of cooling catalyst from the hot stripper. Direct contact heat exchange of relatively hot catalyst in the stripper with a source of relatively cool catalyst provides an efficient and compact method of cooling the hot catalyst from the stripper upstream of the regeneration zone.

The catalyst for direct contact cooling is preferably also taken from the regenerator, although it must be passed through at least one stage of catalyst cooling before being added to the stripping zone.

The process and apparatus of the present invention may be easily added to existing FCC units. Most existing stripper designs, usually with no or only minor modifications, can accommodate the slight increases in mass flow through the stripper caused by direct contact heating of catalyst. This is because FCC units must have stripping zones which will accommodate greatly varying flows, because quite different catalyst to oil ratios are frequently needed to accommodate changes of catalyst activity, reactor temperature required, or changes in feed composition affecting crackability or of regenerator temperature.

To illustrate, most existing FCC unit strippers are designed to operate with up to a 5:1 CAT:OIL ratio. When heavier feeds cause the regenerator temperature to increase, or complete CO combustion in the regenerator makes for hotter catalyst, the reactor does not require nearly as much catalyst circulation to achieve the same top temperature. There is therefor considerable excess capacity in the stripping section when the unit is operating at a CAT:OIL ratio of 3:1.

Assuming that the catalyst stripper can accommodate only a 20% increase in catalyst flow, the following change in stripper

—22—

temperature can be achieved by adding 20% extra hot, regenerated catalyst to the stripper.

ILLUSTRATIVE EMBODIMENT – HEATING

BASIS: Riser top temperature = 538°C (1000°F),
regenerated catalyst temperature = 732°C (1350°F), constant heat capacity assumed, cooling due to stripping steam ignored, as is heat loss due to radiation, etc. Catalyst flow (spent catalyst from stripper) is assumed to be 100 kg/sec (this corresponds to a modest size commercial FCC unit, with a roughly 19,000 BPD oil feed, and a 3:1 Cat:oil ratio.)

IN: 100 kg/s @ 538°C (1000°F)

ADD: 20 kg/s @ 732°C (1350°F)

OUT: 120 kg/s @ 570°C (1058°F).

An increase in catalyst temperature of over 28°C (50°F) will significantly increase the effectiveness of the catalyst stripper.

ILLUSTRATIVE EMBODIMENT – COOLING

BASIS: Use of an external heat exchanger to cool 30 kg/s of hot regenerated catalyst from 732°C to 399°C (1350°F to 750°F). This amount of cooling is readily achievable as there are so many fluid streams circulating around a typical FCC unit with temperatures ranging from ambient to a few hundred °F. Because of the large temperature differential available for heat transfer, a fairly small heat exchanger may be used to achieve catalyst cooling.

IN: 120 kg/s @ 570°C (1058°F)

ADD: 30 kg/s @ 399°C (750°F)

OUT: 150 kg/s @ 536°C (996.4°F)

The traffic through the stripper need only be increased by 20 %, the amount of hot catalyst added. The cooled catalyst can be added at the base of the stripper, or even downstream of

—23—

the stripper, with the cooled and stripped catalyst simple mixing in the transfer line going to the regenerator.

ESTIMATED BENEFITS

5   By operating in this way, significantly enhanced stripping of spent catalyst can be achieved. The coke composition of a typical spent FCC catalyst is reported below, followed by the composition of the same catalyst after conventional stripping, and after the stripping process of the invention.

10   There will be significant reductions in the Wt % coke on catalyst to the regenerator, and in Wt % H in the coke on spent catalyst, as compared to prior art cool stripping process, without increasing the temperature of the stripped catalyst to the regenerator. There will also be a reduction in the % S and %

15   N on stripped catalyst of the invention, and a marked reduction in the temperature rise experienced by the stripped catalyst during the start of the regeneration process, e.g. exiting the riser mixer. The steaming severity of the stripping/regeneration process of the invention will be much less than that of the prior

20   art.

Wt % coke refers to everything deposited on the catalyst to make it spent. It includes sulfur and nitrogen compounds, strippable hydrocarbons, catalytic coke, etc.

Wt % hydrogen in coke refers to the amount of hydrogen

25   that is present in the coke. Most of the hydrogen comes from entrained hydrocarbons or unstripped, adsorbed hydrocarbons. It is a measure of stripping efficiency, and also a indicator of how much water of combustion will be formed upon burning the coke. To a lesser extent, it is an indicator of the extremely high,

30   transient surface temperatures experienced by the catalyst during the start of regeneration. The hydrogen rich materials burn

—24—

rapidly, and are believed to produce large, localized hot spots on the surface of the catalyst.

% S removed refers to all sulfur containing compounds on the spent catalyst and the extent to which these material are
5 rejected in the stripper rather than sent to the regenerator to form SO_x. % N is a similar measure for nitrogen.

The temperature of the catalyst at the riser mixer outlet refers to the measured bulk temperature at the end of a conventional riser mixer as shown in the drawing. The present
10 invention is not limited to use of a riser mixer, but the riser mixer outlet temperature is one of the most sensitive observation points in the regenerator. The process of the present invention has a much smaller rise in temperature through the riser mixer for several reasons. First, there is dilution of spent catalyst
15 with 50 % of regenerated catalyst. This dilution effect aids greatly in damping temperature increases. The second effect is the drastically reduced concentration of strippable hydrocarbons in the process of the present invention. These hydrocarbons burn quickly, and if roughly half of them can be eliminated from the
20 spent catalyst the temperature rise is limited, because the catalytic coke on the catalyst does not burn so quickly.

The reduced surface temperatures are hard to measure. There is no good way known to measure surface temperatures in an FCC, but the results of extremely high surface temperatures have
25 been noted by FCC researchers observing metal migration on FCC catalyst that could only occur at extremely high surface temperatures.

STEAMING FACTOR

The steaming factor, SF, is a way to measure the amount
30 of deactivation that occurs in any part of the FCC process. The base case, or a steaming factor of 1.0, is the amount of catalyst deactivation that occurs in a conventional FCC regenerator

operating at a temperature of 704°C (1300°F), with a catalyst residence time of 4 minutes, in a regenerator with a steam partial pressure of 41 kPa (6.0 psia).

Steaming factor is a linear function of residence time. If a regenerator operates as above, but the catalyst residence time is 8 minutes, then the SF is 2.

Steaming factor is roughly linear with steam partial pressure. SF roughly doubles, or halves, with every change of roughly 25 F. It may be calculated more exactly using the following equation:

$$SF = \frac{(time * PH2O \, EXP(-4500/RT))}{((4) * (6) \, EXP(-4500/(R * 977)))}$$

For a portion of the FCC process operated at 649°C, for a residence time of 2 minutes, and at a steam partial pressure of 172 kPa (10psi), the SF is 0.21.

For an FCC process unit operation at 760°C, steam partial pressure of 110 kPa (PH_2O of 1.0), and a residence time of 4 minutes, the SF is 0.59.

Mathematically, it is calculated using the same temperature effects used for Visbreaking (Base temp. of (427°C)), adjusted for seconds of residence time, and based on a linear extrapolation of steam partial pressure. If FCC catalyst spends 1.0 second at 427°C (800°F) under 101 kPa (1.0 atm) steam partial pressure, then the steaming factor is 1.0 s. Reducing the steam partial pressure to 50.5 kPa (1/2 atm.) would reduce the steaming factor to 0.5 s. Increasing the residence time to 10 seconds, at 760°C (800°F), at 50.5 kPa (0.5 atm.) steam partial pressure would give a steaming factor of 5.0 s. The steaming factor is based on bulk temperatures, so it probably understates the importance of the present invention in reducing the amount of damage done to FCC catalyst by steaming in the regenerator.

The deactivation of FCC catalyst in the unit is of course not just dependent on steaming in the riser mixer in the

—26—

regenerator, but on steaming in every part of the unit, including the steam stripper, deactivation due to metals deposition, etc.

CATALYST REGENERATION

The invention can benefit FCC units using any type of
5   regenerator, ranging from single dense bed regenerators to the more modern, high efficiency design shown in the Figure.

Single, dense phase fluidized bed regenerators can be used, but are not preferred. These generally operate with spent catalyst and combustion air added to a dense phase fluidized bed
10   in a large vessel. There is a relatively sharp demarcation between the dense phase and a dilute phase above it. Hot regenerated catalyst is withdrawn from the dense bed for reuse in the catalytic cracking process, and for use in the hot stripper of the present invention.

15   High efficiency regenerators, preferably as shown and described in the Figure, are the preferred catalyst regenerators for use in the practice of the present invention.

FCC REGENERATOR CONDITIONS

The temperatures, pressures, oxygen flow rates, etc., are
20   within the broad ranges of those heretofore found suitable for FCC regenerators, especially those operating with substantially complete combustion of CO to CO_2 within the regeneration zone. Suitable and preferred operating conditions are:

	Broad	Preferred
25 Temperature, °C (°F)	593-927°C (1100-1700°F)	621-760°C (1150-1400°F)
Catalyst Residence Time, Seconds	60-3600	120-600
30 Pressure, kPa (atmospheres)	101-1010 (1-10)	202-505 (2-5)
% Stoichiometric O_2	100-120	100-105

—27—

CO COMBUSTION PROMOTER

Use of a CO combustion promoter in the regenerator or combustion zone is not essential for the practice of the present invention, however, it is preferred. These materials are
5 well-known.

U.S. 4,072,600 and U.S. 4,235,754 disclose operation of an FCC regenerator with minute quantities of a CO combustion promoter. From 0.01 to 100 ppm Pt metal or enough other metal to give the same CO oxidation, may be used with good results. Very
10 good results are obtained with as little as 0.1 to 10 wt. ppm platinum present on the catalyst in the unit. In swirl type regenerators, operation with 1 to 7 ppm Pt commonly occurs. Pt can be replaced by other metals, but usually more metal is then required. An amount of promoter which would give a CO oxidation
15 activity equal to 0.3 to 3 wt. ppm of platinum is preferred.

Conventionally, refiners add CO combustion promoter to promote total or partial combustion of CO to CO_2 within the FCC regenerator. More CO combustion promoter can be added without undue bad effect – the primary one being the waste of adding more
20 CO combustion promoter than is needed to burn all the CO.

The present invention can operate with extremely small levels of CO combustion promoter while still achieving relatively complete CO combustion because the heavy, resid feed will usually deposit large amounts of coke on the catalyst, and give extremely
25 high regenerator temperatures. The high efficiency regenerator design is especially good at achieving complete CO combustion in the dilute phase transport riser, even without any CO combustion promoter present, provided sufficient hot, regenerated catalyst is recycled from the second dense bed to the coke combustor.
30 Catalyst recycle to the coke combustor promotes the high temperatures needed for rapid coke combustion in the coke combustor and for dilute phase CO combustion in the dilute phase transport riser.

—28—

        Usually it will be preferred to operate with much higher
levels of CO combustion promoter when either partial CO
combustion is sought, or when more than 5-10 % of the coke
combustion is shifted to the second dense bed. More CO
5   combustion promoter is needed because catalysis, rather than high
temperature, is being relied on for smooth operation.

        This concept advances the development of a heavy oil
(residual oil) catalytic cracker and high temperature cracking
unit for conventional gas oils. The process combines the control
10 of catalyst deactivation with controlled catalyst
carbon-contamination level and control of temperature levels in
the stripper and regenerator.

        The hot stripper temperature controls the amount of
carbon removed from the catalyst in the hot stripper.
15 Accordingly, the hot stripper controls the amount of carbon (and
hydrogen, sulfur) remaining on the catalyst to the regenerator.
This residual carbon level controls the temperature rise between
the reactor stripper and the regenerator. The hot stripper also
controls the hydrogen content of the spent catalyst sent to the
20 regenerator as a function of residual carbon. Thus, the hot
stripper controls the temperature and amount of hydrothermal
deactivation of catalyst in the regenerator. This concept may be
practiced in a multi-stage, multi-temperature stripper or a
single stage stripper.

25   Employing a hot stripper, to remove carbon on the
catalyst, rather than a regeneration stage reduces air pollution,
and allows all of the carbon made in the reaction to be burned to
CO_2, if desired.

        The stripped catalyst is cooled by direct contact heat
30 exchange to a desired regenerator inlet temperature. The
catalyst cooler controls regenerator temperature, thereby
allowing the hot stripper to be run at temperatures above the

—29—

riser top temperature, while allowing the regenerator to be run
independently of the stripper.

The present invention strips catalyst at a temperature
higher than the riser exit temperature to separate hydrogen, as
5   molecular hydrogen or hydrocarbons from the coke which adheres to
catalyst. This minimizes catalyst steaming, or hydrothermal
degradation, which typically occurs when hydrogen reacts with
oxygen in the FCC regenerator to form water. The high
temperature stripper (hot stripper) also removes much of the
10   sulfur from coked catalyst as hydrogen sulfide and mercaptans,
which are easy to scrub. In contrast, burning from coked
catalyst in a regenerator produces SO_x in the regenerator flue
gas. The high temperature stripping recovers additional valuable
hydrocarbon products to prevent burning these hydrocarbons in the
15   regenerator. An additional advantage of the high temperature
stripper is that it quickly separates hydrocarbons from catalyst.
If catalyst contacts hydrocarbons for too long a time at a
temperature near or above 538°C (1000°F), then diolefins are
produced which are undesirable for downstream processing, such as
20   alkylation. However, the present invention allows a precisely
controlled, short contact time at 538°C (1000°F) or greater to
produce premium, unleaded gasoline with high selectivity.

The direct contact cooling of stripped catalyst controls
regenerator temperature. This allows the hot stripper to run at
25   a desired temperature to control sulfur and hydrogen without
interfering with a desired regenerator temperature. It is
desired to run the regenerator at least 55°C (100°F) hotter than
the hot stripper. Usually the regenerator should be kept below
871°C (1600°F) to prevent thermal deactivation of the catalyst,
30   although somewhat higher temperatures can be tolerated when a
staged catalyst regeneration is used, with removal of flue gas
intermediate the stages.

—30—

CLAIMS

1. A fluidized catalytic cracking process wherein a
heavy hydrocarbon feed comprising hydrocarbons having a boiling
point above 343°C is catalytically cracked to lighter products
comprising the steps of:

5   a. catalytically cracking the feed in a catalytic
cracking zone operating at catalytic cracking conditions by
contacting the feed with a source of hot regenerated catalyst to
produce a cracking zone effluent mixture having an effluent
temperature and comprising cracked products and spent cracking
10   catalyst containing coke and strippable hydrocarbons;

b. separating the cracking zone effluent mixture
into a cracked product rich vapor phase and a solids rich phase
comprising the spent catalyst and strippable hydrocarbons, the
solids rich phase having a temperature;

15   c. heating the solids rich phase by mixing it with
a source of hot regenerated catalyst having a higher temperature
than the solids rich phase to produce a catalyst mixture
comprising spent and regenerated catalyst having a catalyst
mixture temperature intermediate the solids rich phase
20   temperature and the temperature of the regenerated catalyst;

d. stripping in a primary stripping stage the
catalyst mixture with a stripping gas to remove strippable
compounds from spent catalyst to produce a stripped catalyst
stream;

25   e. cooling a source of hot regenerated catalyst by
passing hot regenerated catalyst through a cooling means to
produce cooled regenerated catalyst;

f. cooling the stripped catalyst stream by direct
contact heat exchange with the cooled regenerated catalyst to
30   produce a cooled, stripped catalyst stream;

—31—

g. regenerating the cooled, stripped catalyst stream by contact with oxygen or an oxygen containing gas in a regenerating means to produce hot regenerated catalyst as a result of combustion of coke on the spent catalyst;

h. recycling to the cracking reaction zone a
5   portion of the hot regenerated catalyst to crack more hydrocarbon feed;

i. recycling to the primary stripping stage a portion of the regenerated catalyst to heat spent catalyst; and

j. recycling to the regenerated catalyst cooling
10   means a portion of the regenerated catalyst to produce cooled regenerated catalyst.

2.   The process of claim 1 wherein the regenerated catalyst cooling means comprises a vessel containing a heat exchanger means, an inlet for hot regenerated catalyst, an outlet for cooled regenerated catalyst, and an inlet for fluidizing gas.

3.   The process of claim 2 wherein the cooled regenerated catalyst is added to the stripped catalyst in the base of the stripping vessel.

4.   The process of claim 2 wherein the cooled regenerated catalyst is added to the stripped catalyst exiting the stripping vessel.

5.   The process of claim 1 wherein the amount of hot regenerated catalyst added is 5 to 50 wt % of the spent catalyst and the temperature of the resulting mixture of hot regenerated and spent catalyst ranges from 28°C above the cracking zone
5   effluent temperature to 833°C.

6.   The process of claim 1 wherein the amount of cooled regenerated catalyst added is 5 to 100 wt % of the spent catalyst.

7.   The process of claim 1 wherein the regenerated catalyst cooler comprises a separate vessel containing a heat exchange means and having an inlet in an upper portion thereof

for hot regenerated catalyst, an inlet in a lower portion thereof for fluidizing gas and an upper outlet for a fluidized mixture of fluidizing gas and cooled regenerated catalyst which flows by gravity to contact the hot stripped catalyst.

8.   The process of claim 1 wherein the catalytic cracking zone comprises a riser reactor.

9.   The process of claim 1 wherein the regenerator comprises:

a riser mixing zone having an inlet at the base thereof for the cooled catalyst mixture and for an oxygen containing gas and an outlet at the top connective with a coke combustion zone;

a coke combustion zone adapted to maintain a fast fluidized bed of catalyst therein, having a catalyst inlet in a lower portion thereof connective with the outlet of the riser mixing zone, an inlet within the fast fluidized bed for additional oxygen or oxygen containing gas, and an outlet in an upper portion thereof connective with a dilute phase transport riser, and wherein at least a portion of the coke on the spent catalyst is burned to form a flue gas comprising CO and CO_2;

a dilute phase transport riser having an inlet in a lower portion thereof connective with the coke combustion zone outlet and an outlet in an upper portion thereof, and wherein at least a portion of the CO in the flue gas is afterburned to CO_2 in the riser to produce at least partially regenerated catalyst which is discharged from the outlet of the dilute phase transport riser into a second dense bed containment vessel;

a dense bed containment vessel adapted to maintain a dense phase fluidized bed of catalyst in a lower portion thereof, having an inlet and separation means connective with the dilute phase transport riser outlet for accepting and separating material discharged from the transport riser into a flue gas rich phase and a catalyst rich phase which is collected as a dense

—33—

phase fluidized bed in a lower portion of the containment vessel, the vessel having regenerated catalyst outlet means connective with the dense phase fluidized bed of catalyst; and

5   catalyst recycle means connective with the catalytic cracking reaction zone and with the primary stage stripping zone.

10.   The process of claim 9 wherein the amount of oxygen or oxygen containing gas added to the riser mixer is limited to limit the temperature rise in the riser mixer and wherein temperatures in the coke combustion zone are increased by
5   recycling of hot regenerated catalyst from the dense bed in the containment vessel to the coke combustion zone to the riser mixer.

11.   The process of claim 1 further characterized in that a CO combustion promoter comprising 0.01 to 50 ppm of platinum group metal or other metal with an equivalent CO oxidation activity, on an elemental metal basis, based on the weight of
5   particles in the regenerator is present on the cracking catalyst.

12.   An apparatus for the fluidized catalytic cracking of a heavy hydrocarbon feed comprising hydrocarbons having a boiling point above 650 F to lighter products by contact the feed with catalytic cracking catalyst comprising:

5   a.   a catalytic cracking reactor means having an inlet connective with the feed and with a source of hot regenerated catalyst and having an outlet for discharging a cracking zone effluent mixture comprising cracked products and spent cracking catalyst containing coke and strippable
10   hydrocarbons;

b.   a separation means connective with the reactor outlet for separating the cracking zone effluent mixture into a cracked product rich vapor phase and a solids rich phase comprising the spent catalyst and strippable hydrocarbons;
15   c.   a hot stripping means having an upper portion and a lower portion and comprising an inlet for a source of hot

—34—

regenerated cracking catalyst in the upper portion thereof, an
inlet for spent catalyst, an inlet for a stripping gas, a
stripping vapor outlet for stripping vapors and a solids outlet
for discharge of hot stripped solids in a lower portion thereof;

5   d. a regenerated catalyst cooling means comprising
a vessel adapted to contain a fluidized bed of catalyst and
having an inlet connective with a source of hot regenerated
catalyst, a heat exchange means immersed at an elevation within
the fluidized bed of catalyst for removal of heat to produce
10 cooled regenerated catalyst, an inlet for a fluidizing gas , and
an outlet for cooled, regenerated catalyst;

         e. a direct contact heat exchange means for contact
and cooling of hot stripped solids with cooled regenerated
catalyst to produce cooled stripped catalyst;

15   f. a catalyst regeneration means having an inlet
connective with the cooled, stripped catalyst, a regeneration gas
inlet, a flue gas outlet, and an outlet for removal of hot
regenerated catalyst; and

         g. catalyst recycle means connective with the
20 catalytic cracking reaction zone, the primary stripping zone, and
the hot regenerated catalyst cooling means.

         13. The apparatus of claim 12 wherein the hot
regenerated catalyst cooler is at an elevation, the hot catalyst
stripper is at an elevation, and both the cooler and stripper are
at substantially the same elevation.

         14. The apparatus of claim 13 wherein the cooler and
stripper are in open communication with a common source of hot
regenerated catalyst, and wherein catalyst flow through the
cooler is only possible when fluidizing gas is added to the
5 catalyst cooler.

FLUE GAS

1/1
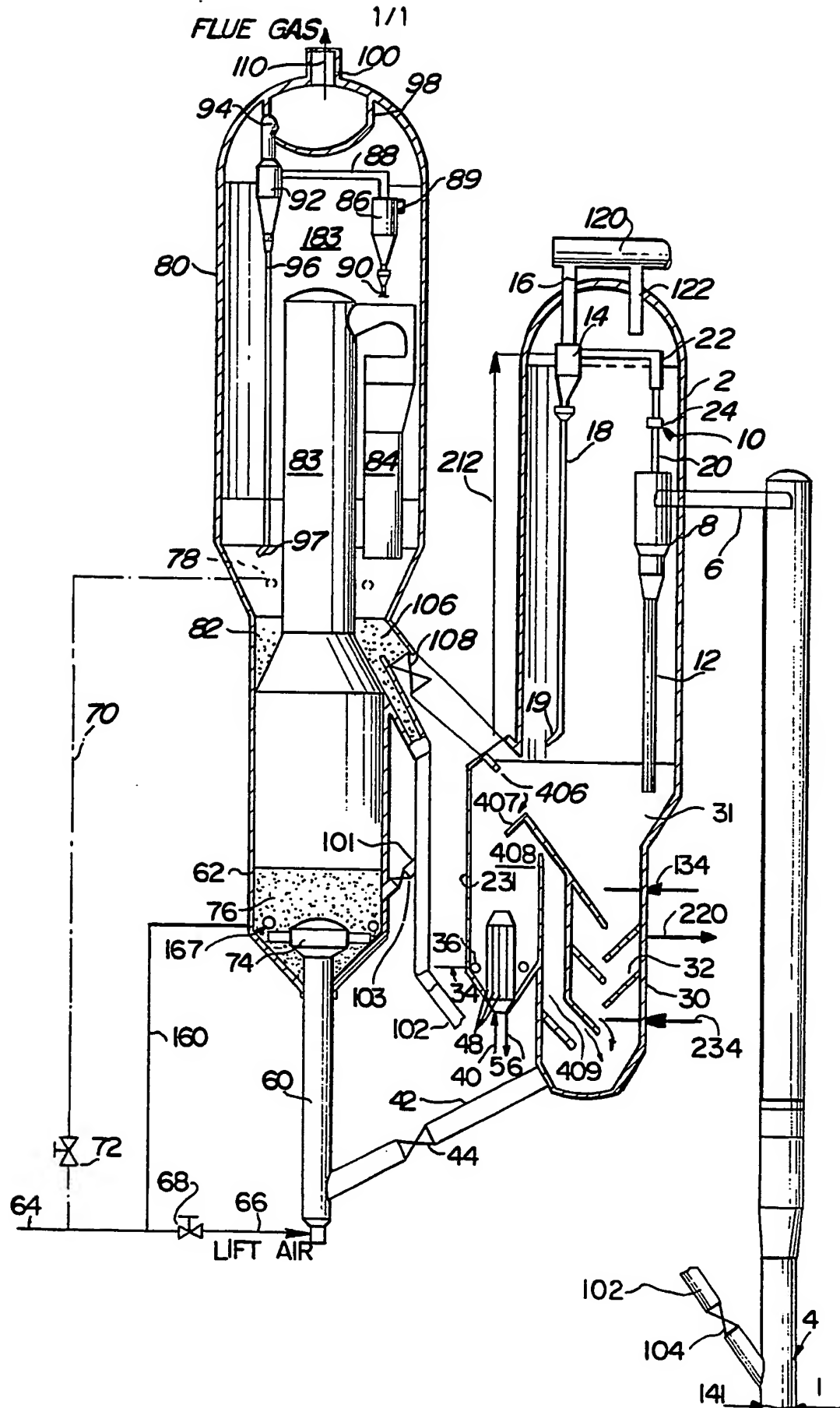
INTERNATIONAL SEARCH REPORT

International Application No **PCT/US 90/01881**

I. CLASSIFICATION F SUBJECT MATTER (if several classification symbols apply, indicate all) *

According to International Patent Classification (IPC) or to both National Classification and IPC

IPC^5: C 10 G 11/18

II. FIELDS SEARCHED

Minimum Documentation Searched [7]	
Classification System	Classification Symbols
IPC^5	C 10 G

Documentation Searched other than Minimum Documentation to the Extent that such Documents are Included in the Fields Searched [8]

III. DOCUMENTS CONSIDERED TO BE RELEVANT [9]

Category *	Citation of Document, [11] with Indication, where appropriate, of the relevant passages [12]	Relevant to Claim No. [13]
X	EP, A, 0234924 (ENGELHARD CORP.) 2 September 1987 see figure 1; claim 1; page 11	1,2,4-6,8,9, 12-14
Y	--	3,7,10,11
Y	EP, A, 0236609 (U.O.P.) 16 September 1987 see figure 1; claims	3,7,10
	--	
Y	US, A, 4235754 (CHESTER et al.) 25 November 1980 see abstract (cited in the application)	11
	--	
P,A	US, A, 4820404 (OWEN) 11 April 1989 see figure; claim 1	1
	--	
	./.	

* Special categories of cited documents: [10]

"A" document defining the general state of the art which is not considered to be of particular relevance

"E" earlier document but published on or after the international filing date

"L" document which may throw doubts on priority claim(s) or which is cited to establish the publication date of another citation or other special reason (as specified)

"O" document referring to an oral disclosure, use, exhibition or other means

"P" document published prior to the international filing date but later than the priority date claimed

"T" later document published after the international filing date or priority date and not in conflict with the application but cited to understand the principle or theory underlying the invention

"X" document of particular relevance; the claimed invention cannot be considered novel or cannot be considered to involve an inventive step

"Y" document of particular relevance; the claimed invention cannot be considered to involve an inventive step when the document is combined with one or more other such documents, such combination being obvious to a person skilled in the art.

"&" document member of the same patent family

IV. CERTIFICATION

Date of the Actual Completion of the International Search	Date of Mailing of this International Search Report
28th June 1990	20. 07. 90

International Searching Authority	Signature of Authorized Officer
EUROPEAN PATENT OFFICE	M. PEIS

Form PCT/ISA/210 (second sheet) (January 1985)

III. DOCUMENTS CONSIDERED TO BE RELEVANT (CONTINUED FROM THE SECOND SHEET)

Category*	Citation of Document, 11 with indication, where appropriate, of the relevant passages	Relevant to Claim No
A	US, A, 4353812 (LOMASS et al.) 12 October 1982 see figure (cited in the application) --------	1

This annex lists the patent family members relating to the patent documents cited in the above-mentioned international search report.
The members are as contained in the European Patent Office EDP file on 17/07/90
The European Patent Office is in no way liable for these particulars which are merely given for the purpose of information.

Patent document cited in search report	Publication date	Patent family member(s)		Publication date
EP-A- 0234924	02-09-87	None		
EP-A- 0236609	16-09-87	US-A-	4578366	25-03-86
US-A- 4235754	25-11-80	None		
US-A- 4820404	11-04-89	None		
US-A- 4353812	12-10-82	US-A-	4424192	03-01-84
		US-A-	4439533	27-03-84
		US-A-	4434245	28-02-84
		US-A-	4483276	20-11-84

EPO FORM P0479